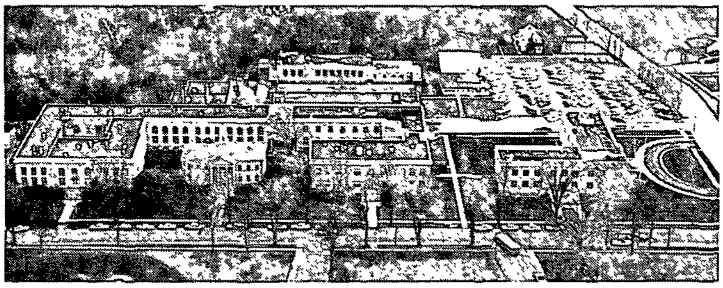

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MAPPS FLUIDIZED BED CALCINATION OF LIME MUD: MODELING FOR DESIGN

DAVID T. CLAY AND NANCY J. SELL

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MAPPS Fluidized Bed Calcination of Lime Mud:
Modeling for Design

David T. Clay

The Institute of Paper Chemistry

Appleton, Wisconsin

and

Nancy J. Sell

University of Wisconsin-Green Bay

Green Bay, Wisconsin

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ABSTRACT

The Institute of Paper Chemistry's MAPPS program was used to model the operation and design of a fluidized bed calciner. In addition to the mass and energy balances, the model can be used to calculate the parameters such as minimum and maximum fluidization velocities, air compressor power requirements and bed size.

INTRODUCTION

A fluidized bed is created when a fluid, often a gas, is passed upward through a bed of relatively fine particles. At low flow rates, the particles remain stationary while the fluid percolates through the voids. At higher flow rates, the particles move apart and become agitated. Particle motion is initially restricted to a small region, but as the flow rate is further increased, the particles will become suspended in the upward flowing fluid. At the point where the weight of the particle is just balanced by the frictional force between the particle and the fluid, the bed is considered to be at minimum fluidization. As the flow rate is increased even further, bubbles will form, and then channeling and slugging will occur.

Fluidized beds can be used to replace the kiln for lime mud calcination in the kraft pulping process. Fluidized bed calciners are commercially available and have been manufactured by companies such as Dorr-Oliver since the early 1960s (1).

A typical fluidized bed calciner is illustrated in Fig. 1. There are two beds: an upper calcining bed and a lower cooler bed. Recycled lime mud particles themselves form these fluidizing beds. The fuel is often oil or possibly natural gas.

A fluidized bed calciner was modeled and programmed into the MAPPS (Modular Analysis of Pulp and Paper Systems) computer simulation system developed by The Institute of Paper Chemistry. The key to how MAPPS functions is the word "modular." The overall flowsheet and the individual components are represented in terms of specific operations and/or specific processes. The calciner model, therefore, has been developed into one of a number of modules which can be combined to simulate an entire kraft recovery system.

In addition to mass and energy balances for the calcination process, the calciner module can be used to estimate engineering design parameters such as fluidizing gas compressor power, bed size, and fuel requirements. Sensitivity tests were made relating these design variables. The model was validated by comparing the results with design parameters from the literature and those supplied by Dorr-Oliver.

(Fig. 1 here)

THE CALCINING SYSTEM

In the kraft pulping process, causticizing and calcining together convert recovered green liquor, primarily sodium carbonate and sodium sulfide, into white liquor, sodium hydroxide, and sodium sulfide. The causticizing reaction converts sodium carbonate and calcium hydroxide into calcium carbonate and sodium hydroxide. The calcining reaction permits recovery of the calcium hydroxide by decomposing the calcium carbonate to calcium oxide (lime) and carbon dioxide. The lime can then be slaked, regenerating the calcium hydroxide. The relevant equations are:

$$Na_2CO_3 + Ca(OH)_2 \longrightarrow CaCO_3 + 2NaOH$$

$$CaCO_3 + \text{HEAT} \longrightarrow CaO + CO_2$$

$$CaO + H_2O \longrightarrow Ca(OH)_2$$

The calcining reaction is endothermic with a heat of reaction of about 3254 kJ/kg of CaO. As heat is supplied, solid calcium carbonate decomposes to solid calcium oxide and gaseous carbon dioxide. The partial pressure of carbon dioxide reaches 101 kPa at about 900 C. At this temperature, transport of the carbon dioxide out of the particle occurs rapidly if the reaction is conducted at atmospheric pressure.

There are two common calcining systems in use in the pulp and paper industry: rotary kilns and fluidized bed calciners. The overall mass and energy balances are similar for both systems. The fluid bed systems report advantages including higher fuel efficiency, smaller space requirements, a more uniformly reactive product, and more rapid startup and shutdown. However, the systems are very sensitive to the sodium content in the lime mud, the optimum being between 0.1-0.4%. Too high a Na concentration will give oversize, unreactive pellets, while too low a concentration will lead to high dust carryover (2).

A typical fluidized bed calcining system is illustrated in Fig. 2. Prior to being fed to the calcining system, the lime mud is filtered while being sprayed by hot wash water for sodium control. The filter cake, approximately 65% solids, is mixed in a paddle mixer with recycled dry mud from the cyclone system. The combined product, at about 85% solids, is taken to a cage mill for further drying and disintegration to a fine powder. The mixture is then returned to the cyclones. The fines exit with the flue gas to a scrubber system. Most of the dry calcium carbonate collects in the cyclone. Some is sent back to the paddle mixers while the remainder is pneumatically conveyed to the fluidized bed for calcination. The fine calcium carbonate particles react quickly at about 900°C to form lime, which then adheres to the other pellets in the bed, causing them to grow. The pellets are randomly removed from the calcining bed as they overflow through a stand pipe downward to the cooling bed. The incoming fluidizing air reduces the particle temperature to a level above 260°C in the cooling bed. From the cooling bed the pellets are removed to a product elevator, then to storage.

(Fig. 2 here)

The MAPPS version of the fluidized bed calcining system flowsheet is in Fig. 3. As can be noted, much of the various equipment is simulated by a combination of simple units. For example, the cage and paddle mixers (pug mills) are represented by a mixer, the conveyor system by a mixer and splitter, the cyclones by two separators and a mixer, CaO storage by a heat exchanger, and similar representations. The CONVRT module is a requirement of the simulation program, not representative of any process equipment. It transfers the flow rates of the various components from their location in a gaseous type stream to the proper location in a recovery type stream, so the flue gas can be recycled within the system.

(Fig. 3 here)

THE CALCINER MODEL

There were a large number of approaches which could have been used to develop the equations required. After evaluating the work which had been done by a variety of researchers, we chose to use an approach similar to that used by Kunii and Levenspiel (3) for the design calculations. Not only did their equations appear to be readily adaptable to MAPPS format, but other researchers had found their work to agree best with the experimental results, at least for minimum and maximum superficial velocities (4).

The overall approach taken in the design was the following. The heat requirement is calculated from the mass and energy balance based on the desired product lime flow rate. That fuel requirement is combined with the specified residence time, fluidizing gas velocity, and excess air to calculate the calciner diameter, air compressor power requirements, and related process or equipment parameters.

A variety of empirical correlations specific to calciners was developed by Ducar and Levin (5). Their data were obtained from a literature survey and field trips to four pulp mills and one sewage treatment plant. The systems under consideration ranged in calciner bed diameter from 1.8 to 3.4 meters. These correlations were also included in this module.

Mass and Energy Balances

For the purpose of the overall mass and energy balances, the "black box" approach can be taken for the fluidized bed. Figure 4 illustrates the input and output mass flows.

(Fig. 4 here)

A number of assumptions must be made for this type of calculation. Example values for the parameters listed below are given in Table 1.

- The inlet feed composition and the product lime flow rate must be specified.

- It is necessary to specify the lime conversion efficiency, operating bed temperature, lime exit temperature, incoming air temperature, fuel composition and temperature, outlet temperature of the lime and the flue gas, excess air required, dust loss, and heat loss (as a fraction of total enthalpy).

- It is assumed the flue gas exits at the calciner bed temperature and the lime product exits at the temperature of the cooler bed.

- A fixed amount of fuel can be specified if, for example, it is desired to burn a waste fuel. All of this fuel will be burned, regardless of the specified operating temperature.

- If additional fuel is needed, this amount will be calculated. If a zero flow rate is specified for the fixed fuel, the total fuel requirement will be calculated as "extra fuel." The composition of this extra fuel must be specified as "fuel information." The flow rate specified for the fuel information stream is ignored.

- The lime dust stream is initially calculated separately from the flue gas stream, and then the two are combined. The flow rate of the lime dust stream is always set equal to zero. This combination of streams simulates the entrainment of the dust in the flue gas stream, such as actually occurs. This combination of two stream types requires a redefinition of the gaseous and recovery type streams at run time.

- The fuel is assumed to burn completely. All elemental sulfur, such as that specified in a high sulfur fuel, is assumed to react with the lime to produce calcium sulfate.

Table 1. Parameter values used for the example mass and energy balance listed in Table 2.

Product lime flow	6,483 kg/hr
Inlet composition to fluid bed calciner	Wt. %
$CaCO_3$	79.1
$Ca(OH)_2$	9.0
CaO	6.5
H_2O	5.4
Feed temperature	303°C
Lime conversion efficiency	85%
Fluid bed operating temperature	860°C
Lime exit temperature	343°C
Incoming air temperature	26°C
Fuel temperature	32°C
Excess air	10%
Heat loss	10%
Fuel type	Natural gas
Reference temperature	25°C

The results of a typical mass and energy balance are in Table 2.

Table 2. Results of fluid bed mass and energy balance calculations.
(Conditions are listed in Table 1)

	Mass Flow (kg/hr)	Energy (kJ/hr)
Input		
Fuel	832	46.2×10^6
Calciner feed	11,461	3.71×10^6
Air	15,812	0.02×10^6
TOTAL	28,105	49.9×10^6
Output		
Flue gas + dust	21,622	27.3×10^6
Lime product	6,483	1.79×10^6
Heat of reaction	--	16.2×10^6
Heat loss	--	4.62×10^6
TOTAL	28,105	49.9×10^6

Design Calculations

The mass and energy balances serve as the framework for the calculation of the various design specifications. The following is a summary of some of the relationships used which are specific to fluidization and/or to fluidized beds.

Minimum Fluidization Velocity. The onset of fluidization begins when the drag force due to the upward flowing gas equals the weight of particles. This point is dependent on the bed voidage at that minimum fluidization point, the shape and size of the particles, and the Reynolds number based on the particle diameter. For Reynolds numbers greater than 1000 it is found that the minimum fluidization velocity can be approximated (3):

$$Vmin = (((Dpart**2)*(Dsolid - Dgas)*g)/(24.5*Dgas))**0.5 \qquad (1)$$

where Dpart = average particle diameter

Dgas = fluidizing gas density at actual temperature and pressure

Dsolid = density of particle

g = gravitational constant

As is apparent, an average particle diameter must be assumed for this calculation. In actuality, there will be a distribution of diameters, and the diameters will vary with residence time within the calcining section. The calcination reaction itself would encourage a shrinking of the particle diameter, but the particle agglomeration which occurs due to the stickiness of the sodium salts can overwhelm this effect. The particle density is also an approximate figure, since both a chemical reaction is occurring and the material is drying. The gas density value used is that indicative of conditions at the bottom of the calcining bed.

Maximum Fluidization Velocity. The maximum fluidization velocity is determined as the maximum permissible without significant particle carryover. This upper limit to gas flow rate is approximated by the free-fall velocity of the particles, and is dependent on the drag coefficient, an experimentally determined number. If we assume the Reynolds number for the particle is between 500 and 200,000, the drag coefficient can be approximated as 0.43, hence the terminal velocity for spherical particles is given by (3):

$$Vmax = ((3.1*g*(Dsolid - Dgas)*Dpart)/Dgas)**0.5 \qquad (2)$$

The gas density is now indicative of that at the top of the calcining bed.

As noted, the range of applicability of these two equations is determined by the Reynolds number. If a Reynolds number for a calcining system is calculated, it will likely be lower than the ranges specified, i.e., 450 would be typical. However, the use of equations (1) and (2) is appropriate. Since the minimum velocity is the more important for design considerations, an example of the first equation's applicability will be discussed. Hand-calculation of the minimum velocity using a form of equation (1) with fewer assumptions and hence applicable also at lower Reynolds numbers yielded velocities not significantly different than that given from the equation. The more exact minimum velocity of 1.25 m/sec closely compared to that calculated from equation (1) of 1.27 m/sec at a Reynolds number of 450. Using a form of the equation valid only at lower Reynolds numbers gives significantly different velocities, e.g., 1.90 m/sec compared to 1.25 m/sec.

The desired bed velocity specified by the user must be in the range between the minimum and maximum values or an error message will be given.

Bed Areas and Weights. The gas flow to the calciner is assumed to be that of the flue gas stream, without the dust, and the flow to the cooler is assumed to be the fluidizing gas. The area of each of the beds is then calculated by:

$$\text{Area} = \text{Flow}/(\text{Vbed}*\text{Dgas}) \tag{3}$$

where Vbed = specified superficial velocity through the bed

Dgas = gas density at bottom of bed

The settled height of a bed of fine particles is approximately that at minimum fluidization; little increase in voidage occurs due to the fluidization process initially. Therefore, the bed weights are calculated from the specified heights at minimum fluidization (<u>3</u>). The bed height is assumed to remain constant.

$$Weight = Height*Area*(1-Void)*Dsolid \tag{4}$$

where Void = voidage, typically around 0.5

Air Compressor Power Requirements. There are a variety of air compressor types which could be used: one- or multistage, with isothermal or adiabatic operation assumed. This model is based on a one-stage, adiabatic system. The results of using a two-stage adiabatic or a one-stage isothermal compressor are not significantly different under typical operating conditions.

For the one-stage adiabatic system, the ideal power can be calculated (6):

$$Power = Pl*Qcomp*(Pin/Pl)**Ratio)/Ratio \tag{5}$$

where Pin = the inlet pressure of the fluidizing gas

Pl = gas pressure in the plenum

Qcomp = the volumetric flow rate of fluidizing gas, or mass flow/density

Ratio = (Gamma - 1)/Gamma, and

Gamma = the ratio of the heat capacities, Cp/Cv

SENSITIVITY TESTS

Sensitivity tests were then made for the calciner module itself. The relevant inlet and outlet flows are indicated in Fig.

4. A number of the variables were systematically altered to check the correlation with other parameters. The results can be seen in Fig. 5-9. The "base case" data are summarized in the Appendix.

(Fig. 5-9 here)

Air Compressor Ideal Power *vs.* Feed Rate (Fig. 5)

The function is linearly increasing, as would be expected from Eq. (5). The calciner model requires a constant depth bed, hence the bed area and volumetric flow are also increasing linearly.

Air Compressor Ideal Power *vs.* Excess Air (Fig. 6)

The volumetric flow rate is proportional to the extra air specified, and the power is linearly related to the volumetric flow rate [Eq. (5)]. The slight nonlinearity arises because more air requires more fuel to obtain a given temperature. The additional fuel requirement leads to an additional air requirement to maintain a constant fractional excess.

Fuel *vs.* Excess Air (Fig. 7)

Additional air requires more fuel to heat the gases to the required temperature. The increasing slope is due to heat loss being specified as a fraction of the total enthalpy into the calciner.

Calciner Bed Diameter *vs.* Feed Rate (Fig. 8)

The function is increasing but with a decreasing slope. The feed *vs.* bed area would be linear, but the area is proportional to the diameter squared.

Minimum Fluidization Velocity *vs.* Particle Diameter (Fig. 9)

The minimum velocity is proportional to the square root of the particle diameter, as in Eq. (1).

MODULE VALIDATION

Validation of a simulation is the checking of the model against actual commercial data. The calculated results of this module were thus compared with design values from the literature (5,8,10,11) and supplied by Dorr-Oliver (7).

Figures 10-12 illustrate comparisons between the calculated and reported values for fluidizing gas compressor design power, the inside diameter of the calcining section, and the volumetric flow rate through the air compressor. For these comparisons, an average production rate of 156 t/d of product was chosen. The other calciner specifications chosen were typical for such systems. Most of the actual values used were the same as the base case and are listed in the Appendix. There were some exceptions. The superficial velocities in both the cooler and calciner sections were changed to 0.92 m/s, the particle size was changed to 0.84 mm, and the compressor efficiency was changed to 65%. No attempt was made to exactly match up all conditions between each literature source and the model. Despite this there is good agreement.

(Fig. 10-12 here)

Fluidizing Gas Compressor Power (Fig. 10)

The line represents the models predictions. The commercial data are shown with individual points. The calculated line is the design power, using a 65% compressor efficiency. Since the literature values are typically brake power an additional 15% was added to the calculated operating power. Further, a second 15% was added for design safety. Both of these factors are typical of actual design practice.

Inside Diameter of Calcining Section (Fig. 11)

The calculated curve is slightly below that of literature reported design values. The particle size used for these calculations was 0.84 mm. The resultant calculated velocity was 0.92 m/s. According to Moran (8) 85% of the lime products is above this size. When larger particle sizes were selected the calculated diameters were unreasonably low. The results suggest that use of a fluidizing nominal size instead of a weight nominal size may more accurately represent the phenomena.

Volumetric Flow Rate of Fluidizing Gas at the Blower (Fig. 12)

The flow calculated by the calciner module is the actual volumetric flow exiting the compressor. The commercial values were at standard conditions. The calculated values were, therefore, converted to standard cubic meters per second values for this comparison. The model line is in reasonable agreement with the commercial design data. Differences in excess air between the data and the model may account for most of the variation.

CONCLUSIONS

The fluidized bed calciner module can be used for the calculation of not only mass and energy balances, but also for predicting process engineering design specifications common in commercial use. This latter is an extension of the present functions of MAPPS, but the overall framework of the system is such that it can easily be adapted for this purpose.

ACKNOWLEDGMENTS

We sincerely thank Dorr-Oliver for providing much of the technical data which was essential for the validation of this model. We want to acknowledge the support from member companies of The Institute of Paper Chemistry for portions of this work.

Portions of this work were used by N. J. Sell as partial fulfillment of the requirements for the Master of Science Degree at The Institute of Paper Chemistry.

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APPENDIX

CALCINER BASE CASE PARAMETERS

PROGRAM INPUTS

Calciner feed	11,461 kg/h (5,216 lb/hr)	
Feed temperature	303°C (578°F)	
Feed composition	See Table 1	
Conversion	85%	
Dust loss	10% of product	
Lime temperature	343°C	(650°F)
Flue gas temperature	860°C	(1580°F)
Fluidizing gas temperature	26°C	(80°F)
Excess air	25%	

Heat loss (i.e., let program calculate the required amount) 10% of fuel input enthalpy

Fixed fuel flow rate	0	
Voidage	0.5	
Density of solid	1682 kg/mm^3	(105 lb/ft^3)
Density of fluidizing gas at STP	1.29 kg/mm^3	(0.0807 lb/ft^3)
Superficial velocity through cooler	1.6 m/s	(5 ft/sec)
Superficial velocity through calciners	1.6 m/s	(5 ft/sec)
Particle diameter	1.3 x 10^{-3} m	(0.004 ft)
Calcining pressure	101 kPa	(14.7 psi)
Compressor efficiency	100%	

Height of cooler bed	0.66 m	(2 ft)
Height of calciner bed	2.6 m	(8 ft)
Diameter of orifices in distributor plates		
	1.6×10^{-3} m	(5×10^{-3} ft)

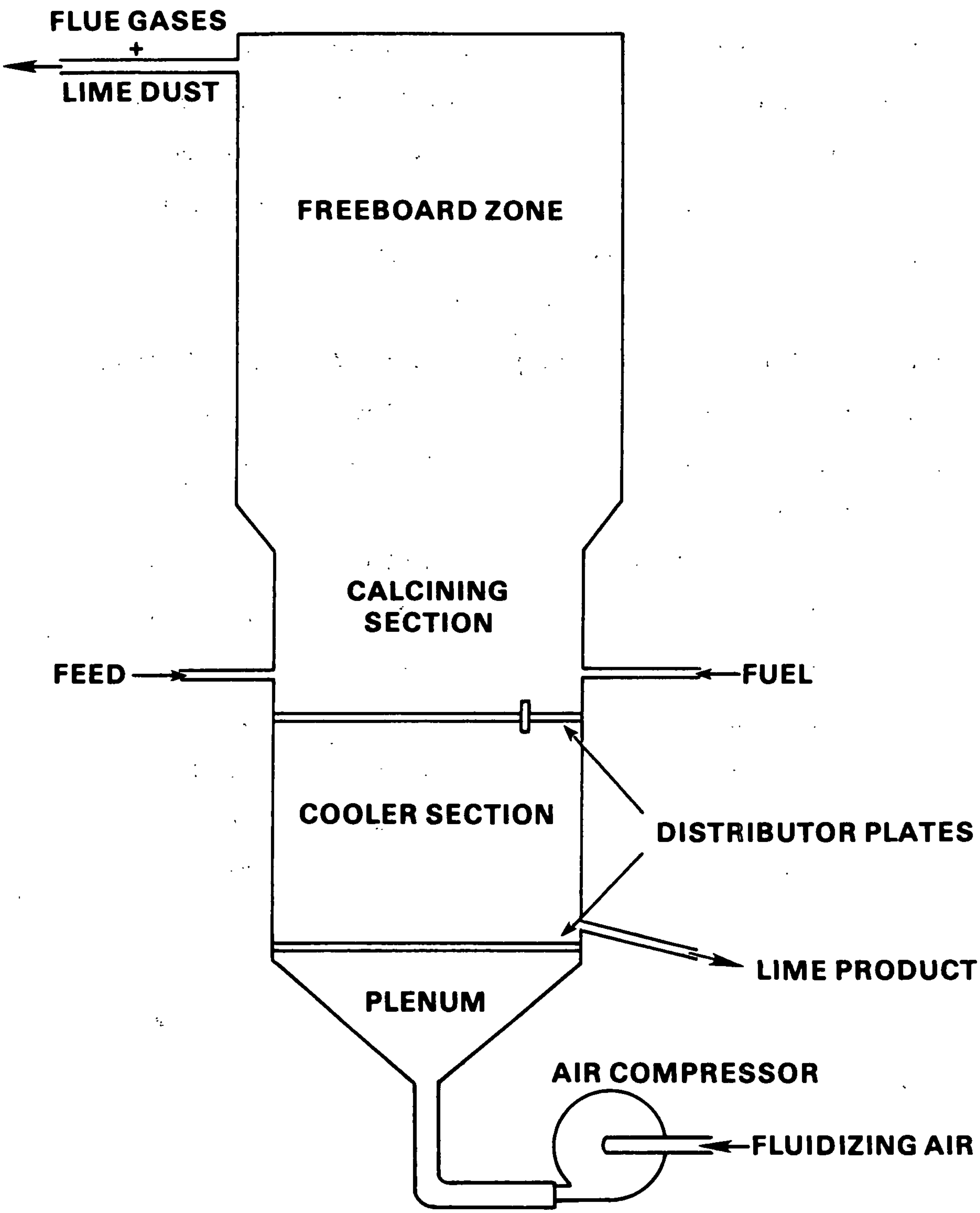
PROGRAM CALCULATED VALUES

Extra fuel required	924 kg/h	(2035 lb/hr)
Adiabatic flame temperature	1730°C	(3146°F)
Cooler bed temperature	343°C	(650°F)
Calciner bed temperature	860°C	(1580°F)
Minimum superficial velocity	0.92 m/s	(3.0 ft/sec)
Maximum superficial velocity	12 m/s	(38 ft/sec)
Volumetric flow rate at conditions		
- cooler	1.22×10^4 m^3/s	(4.7×10^5 ft^3/sec)
- calciner	3.39×10^4 m^3/s	(1.30×10^6 ft^3/sec)
Diameter of cooler bed	1.6 m	(5.3 ft)
Diameter of calciner bed	2.7 m	(8.9 ft)
Area of cooler bed	2.1 m^2	(22.2 ft^2)
Area of calciner bed	5.7 m^2	(61.8 ft^2)
Ideal power	1.25×10^5 J/s	(168 hp)
Pressure drops - cooler bed	5 kPa	(0.73 psi)
- calciner bed	20 kPa	(2.90 psi)
- distributor plate (cooler)	3.4 kPa	(0.50 psi)
- distributor plate (calciner)	3.4 kPa	(0.50 psi)

Weight of calciner bed	1.26×10^4 kg	(2.78×10^4)
Weight of cooler bed	1.14×10^3 kg	(2.51×10^3)
Residence time - calciner	4212 s	(70 min)
- cooler	630 s	(10 min)
No. of plate openings - cooler	2.13×10^4	
- calciner	1.44×10^4	
Volumetric flow rate thru compressor	3.56 m^3/s	(126 ft^3/sec)
Diameter of freeboard zone	4.58 m	(15 ft)
Area of freeboard zone	16.5 m^2	(178 ft^2)
Number of feed points	7	
Height - expanded calciner bed	3.93 m	(12.9 ft)
- plenum	0.9 m	(2.95 ft)
- calciner section	3.2 m	(10.5 ft)
- expansion section	0.15 m	(0.5 ft)
- cooler section	2.6 m	(8.5 ft)
- freeboard zone	4.9 m	(16 ft)
- overall	11.6 m	(38 ft)

Figure 1.

FLUIDIZED BED CALCINER



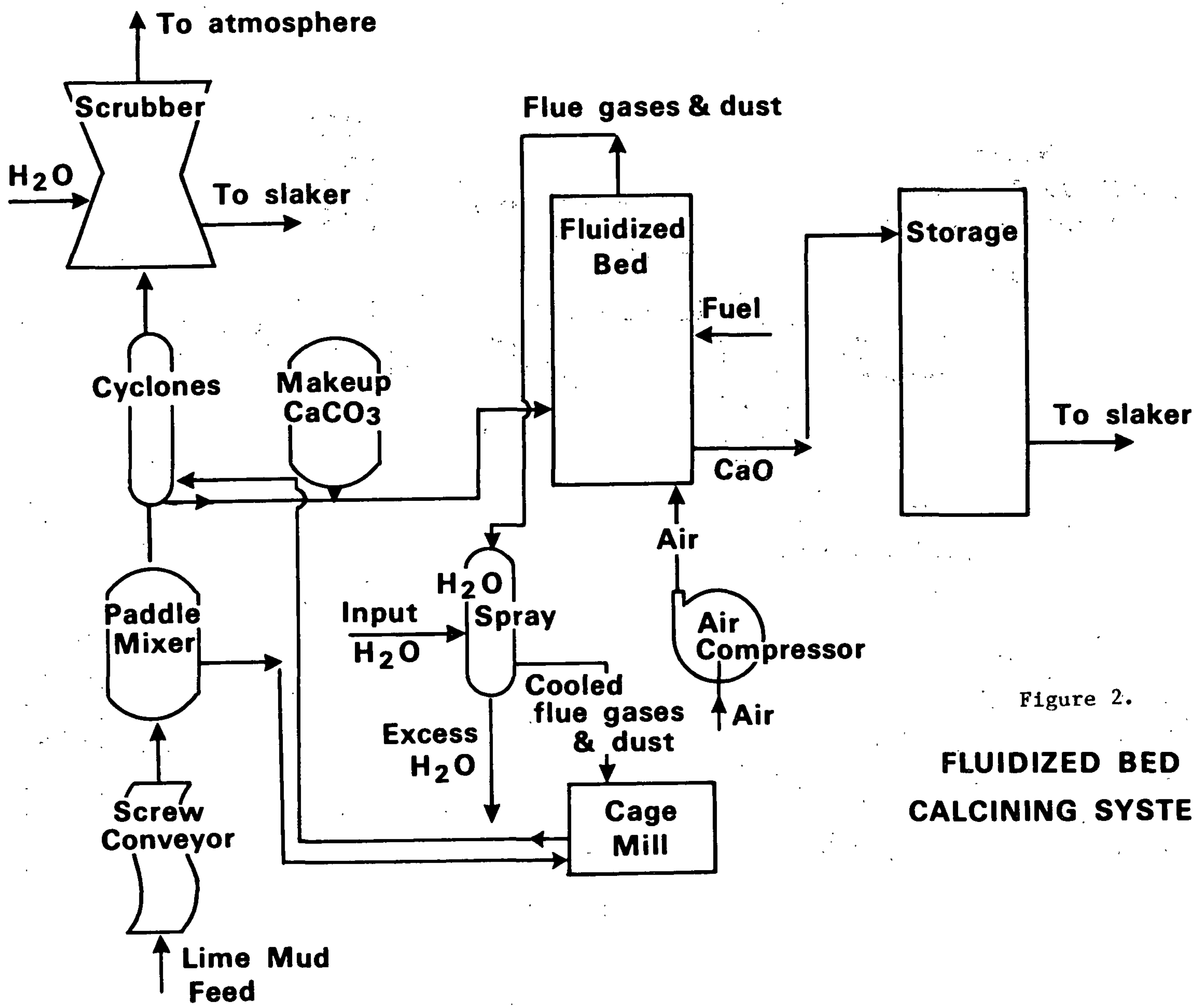


Figure 2.

FLUIDIZED BED CALCINING SYSTEM

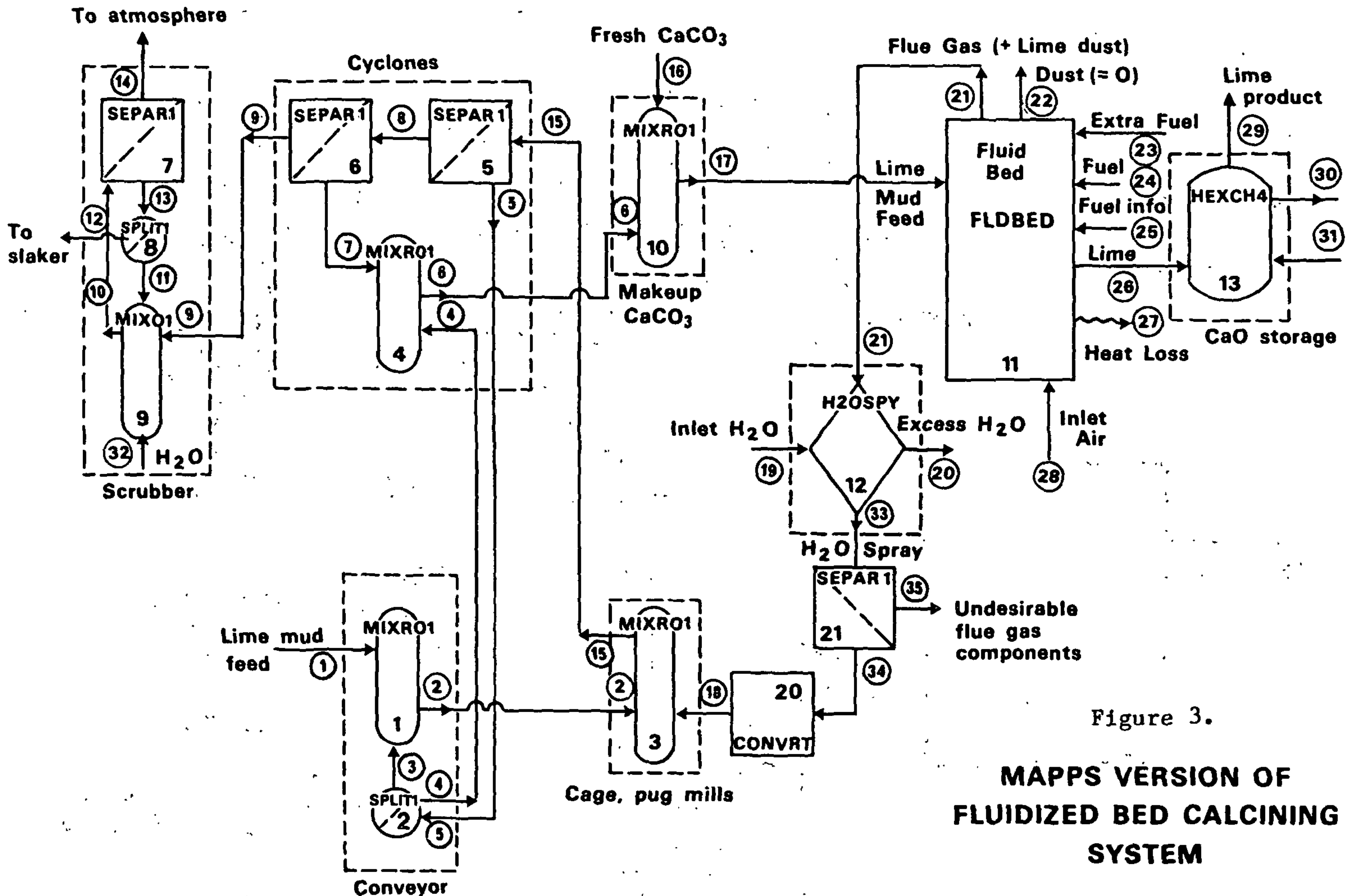
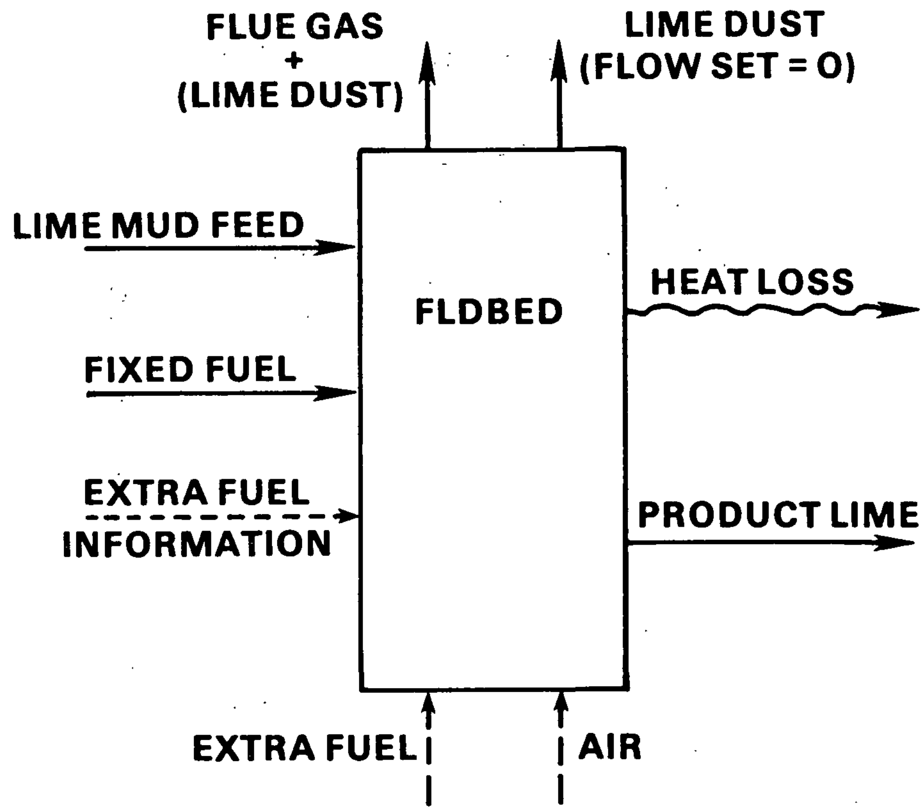


Figure 3.

MAPPS VERSION OF FLUIDIZED BED CALCINING SYSTEM



CALCINER MASS & ENERGY FLOWS

Figure 4.

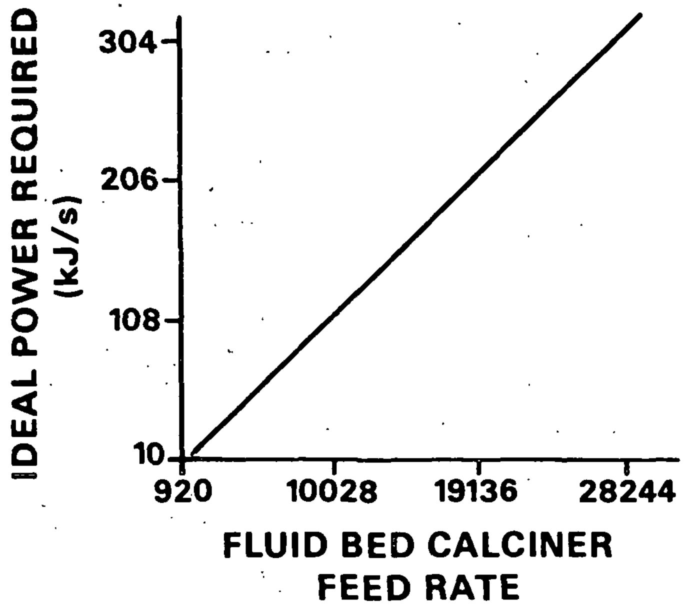

IDEAL POWER REQUIRED
(kJ/s)
304
206
108
10
920
10028
19136
28244
FLUID BED CALCINER
FEED RATE

Figure 5.

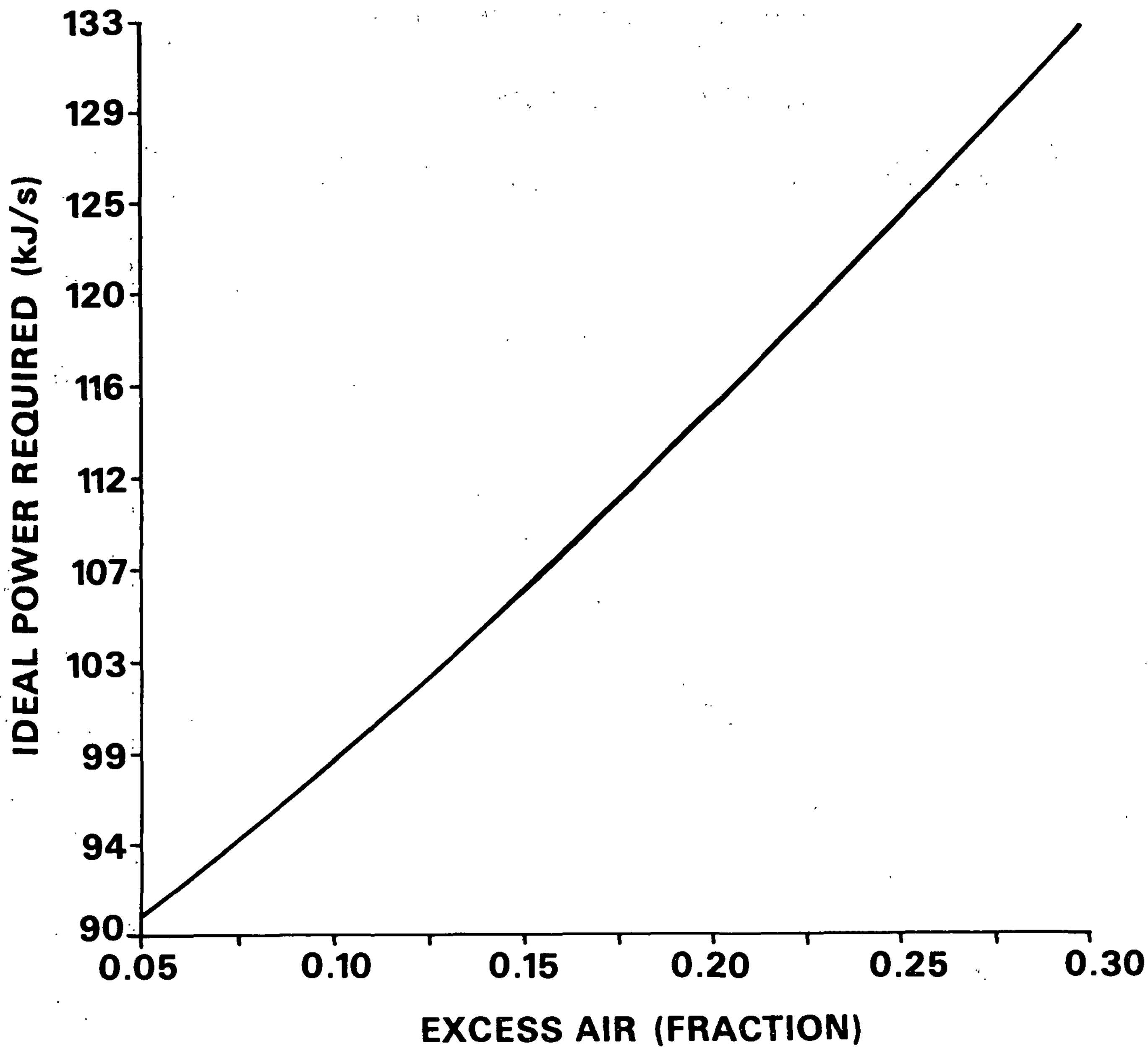


Figure 6. The increase of ideal compressor power with the excess air level.

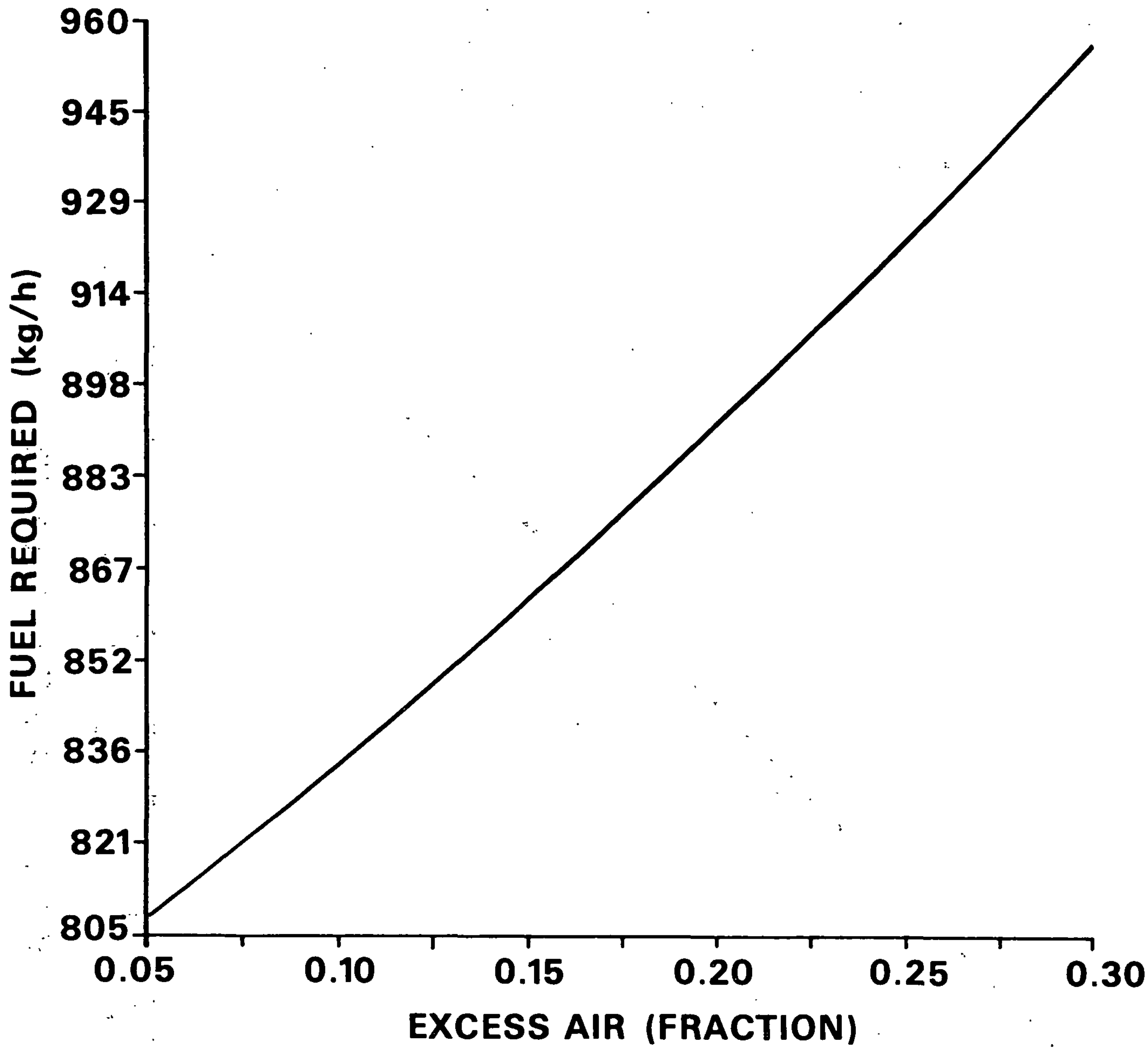


Figure 7. Calciner fuel as a function of excess air.

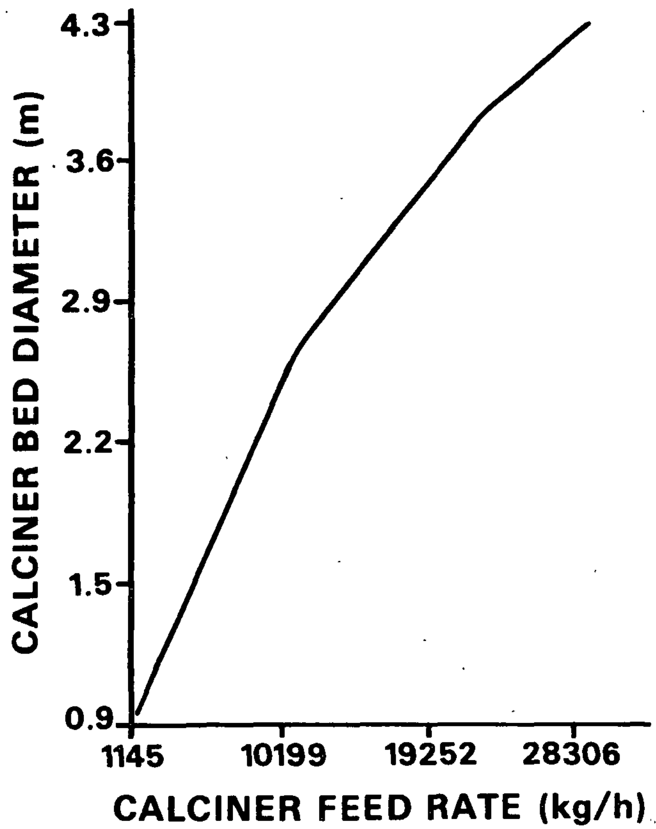


Figure 8. The dependence of calciner inside diameter on feed rate.

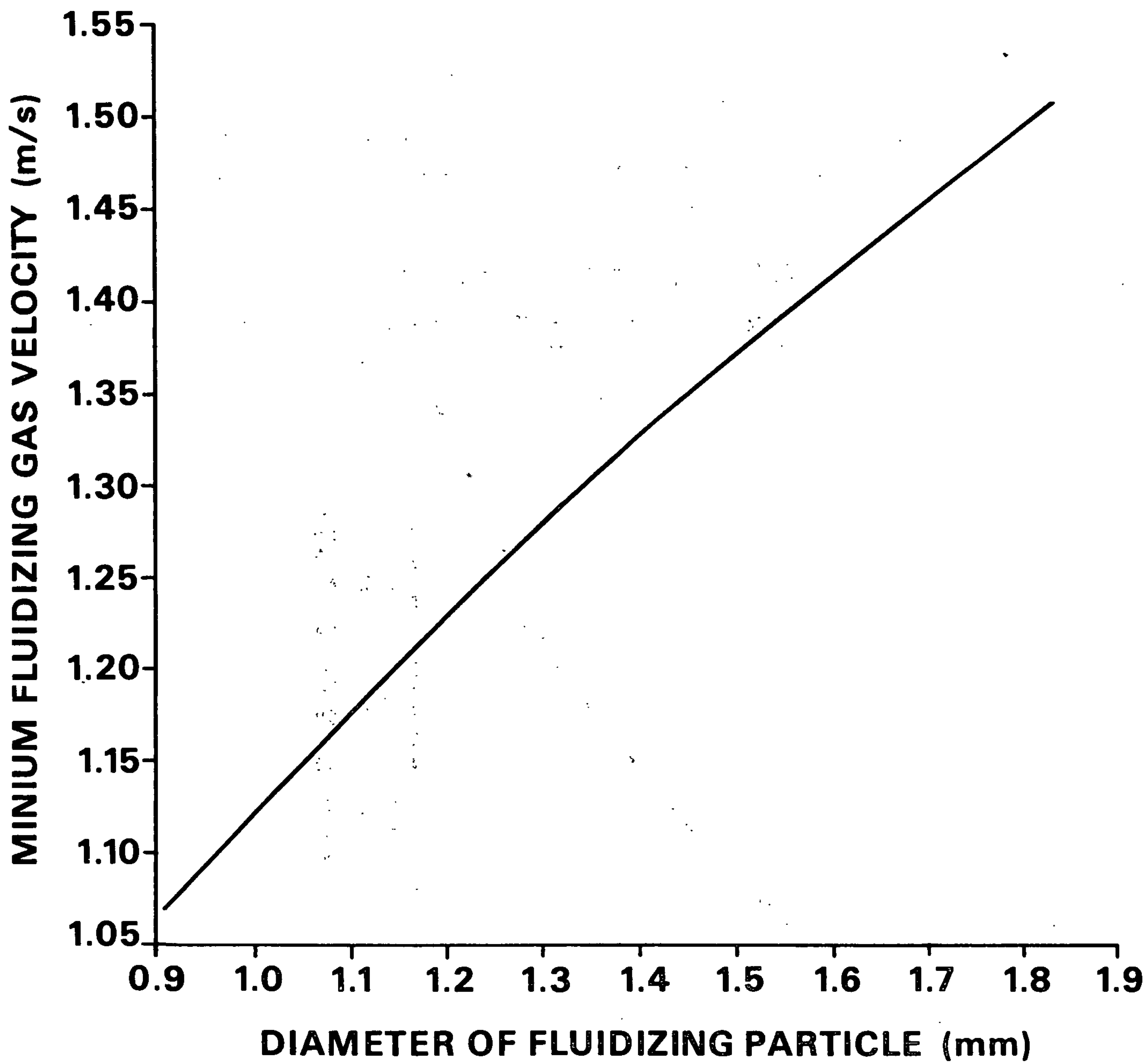


Figure 9. Calculated minimum fluidization velocity as a function of particle size.

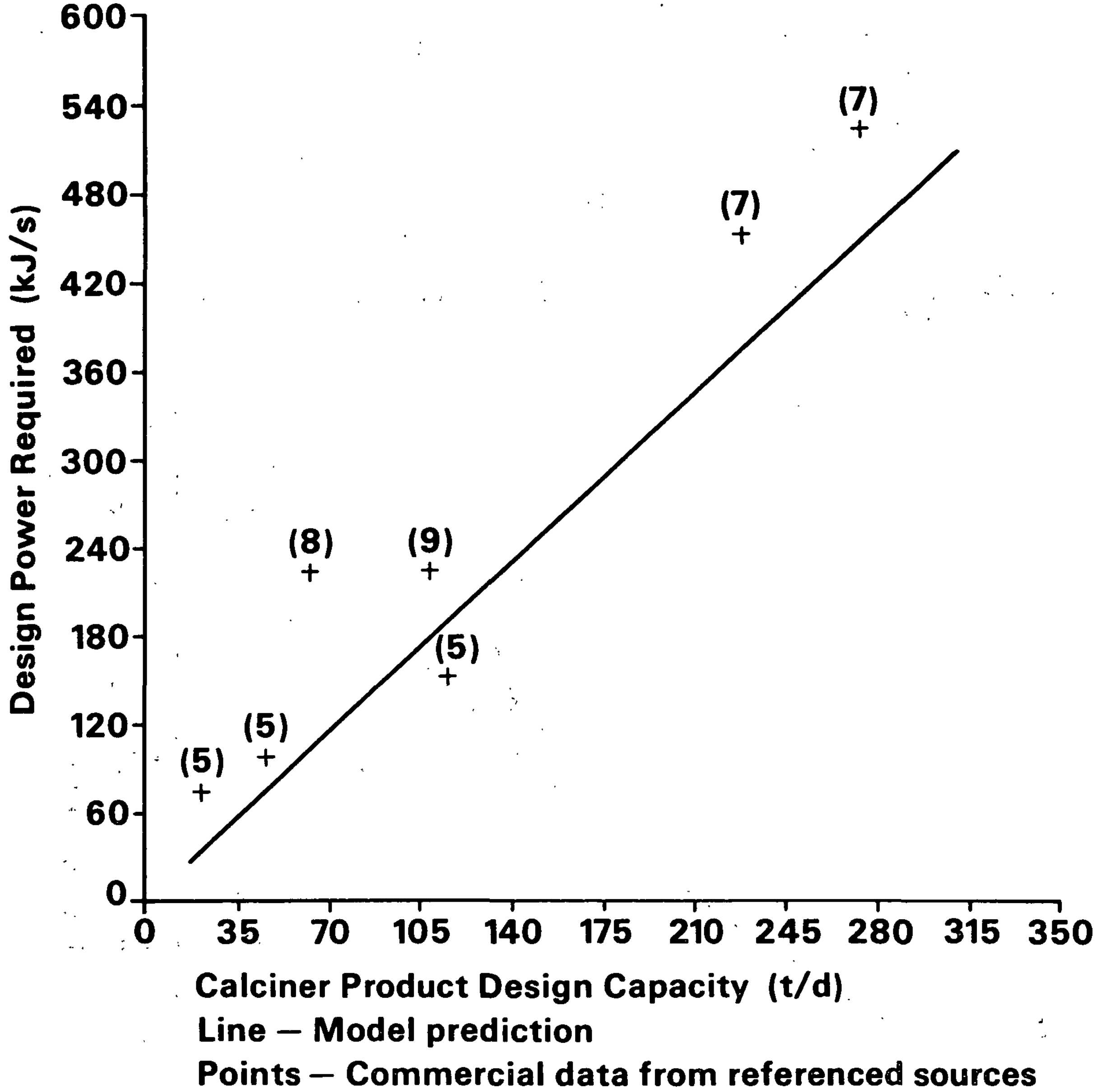


Figure 10. A comparison of literature values with model predictions for design compressor power requirements for a given calciner design capacity. ( ) = reference.

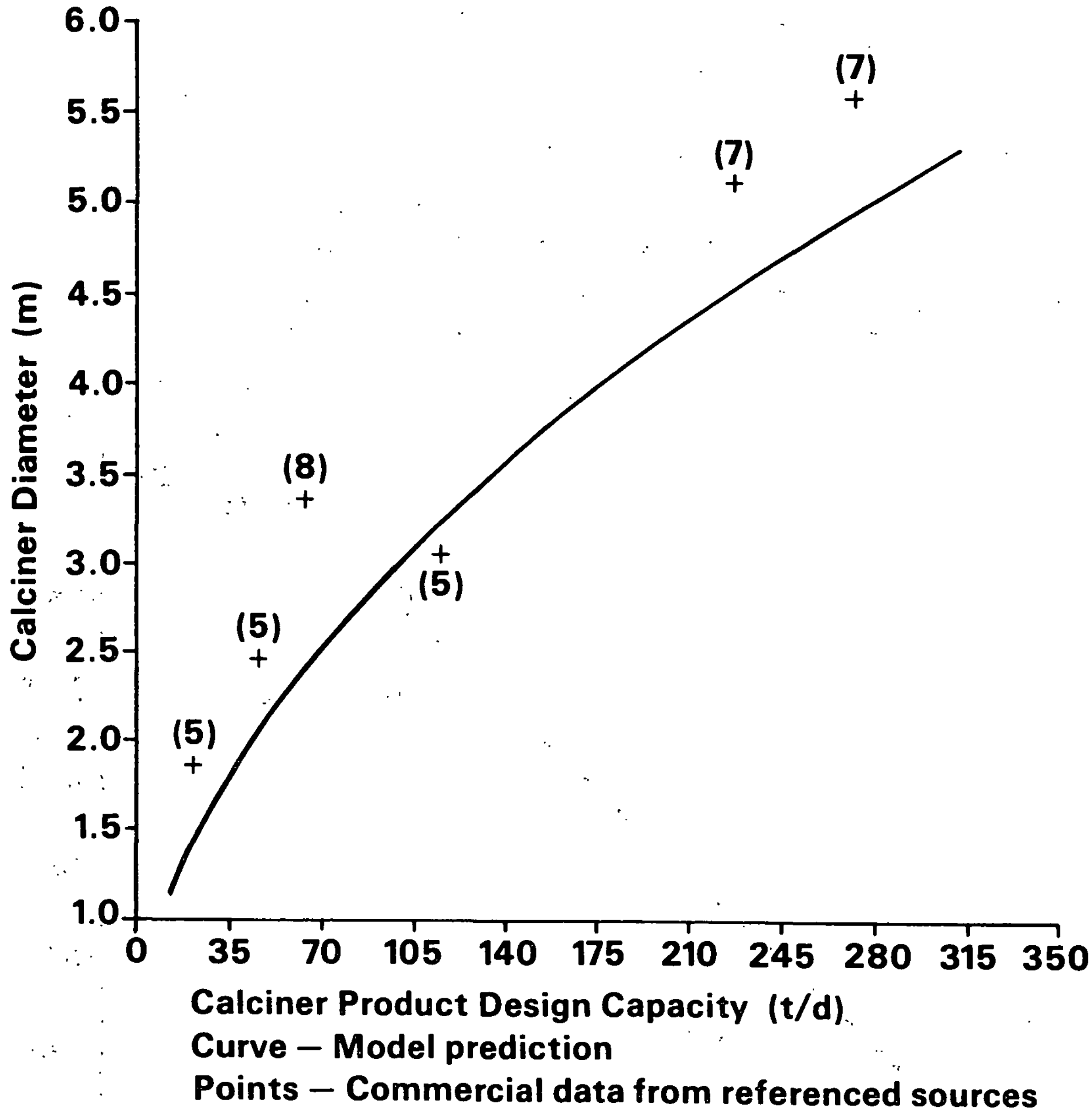


Figure 11. Comparison of literature values with model predictions for calciner inside diameter required for a given calciner capacity. ( ) = reference.

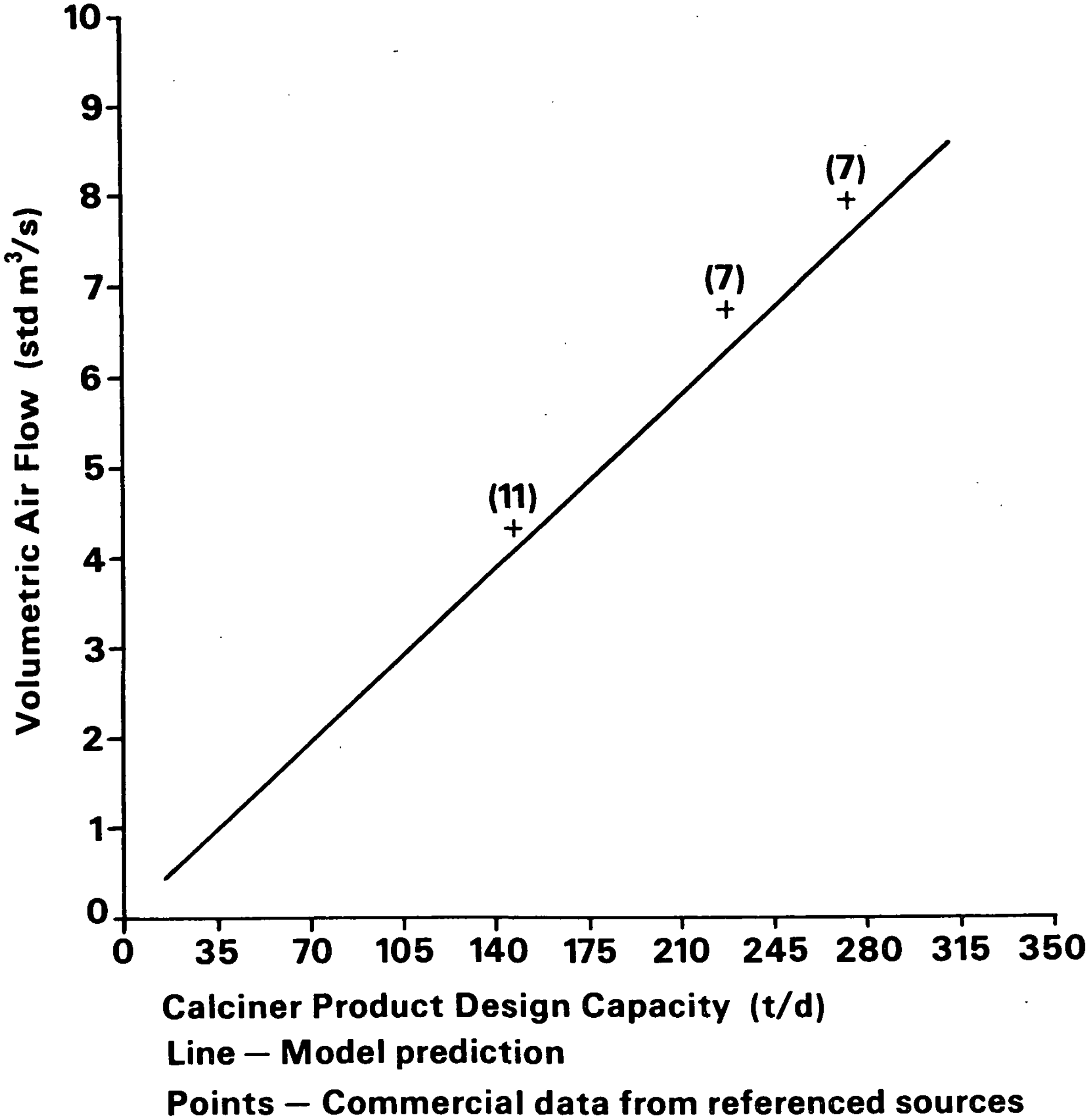


Figure 12. Comparison of literature values with model predictions for compressor flow rates required for a given calciner capacity. ( ) = reference.